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COMPARISON OF BUBBLING AND CIRCULATING FLUIDIZED BED INDUSTRIAL STEAM GENERATION

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ABSTRACT

Atmospheric Fluidized-Bed Combustion (AFBC) for the generation of steam is a viable option for industrial users. This combustion technology has made great advances in general acceptability in recent years. Numerous tests and demonstration projects, including recent commercial units, have proven its technical merits. This paper compares the technical and economic options which face an AFBC project developer.

The generally accepted advantages of AFBC for industrial steam generation are:

- o Environmental acceptability or greater adaptability to existing and proposed future environmental regulations.
- o Greater fuel flexibility - the ability to burn low grade as well as high grade fuels including waste fuels.

Industry has generally accepted the Bubbling Fluidized Bed (BFB) and Circulating Fluidized Bed (CFB) combustors for burning solid fuels to generate steam. This paper considers both dense bed BFB and high velocity lean phase CFB. A technical comparison has been made between the BFB and CFB technologies which evaluates the differences and merits of each. The paper presents capital costs data for a range of unit sizes. Annual O&M costs were also developed, and are summarized on several curves which compare annual total cost per unit size, as well as cost/lb/hr of steam produced.

Numerous issues and options face a potential AFBC owner. The information presented attempts to answer many of these questions which will allow the developer to understand the important decisions which must be made.

INTRODUCTION

Both the BFB and CFB technology should be competitive for industrial steam generation, especially if credit is taken for SO_2 capture and NO_x reduction. The CFB has slightly better performance characteristics for combustion efficiency and limestone utilization for sulfur capture. However, the CFB design does not scale down well. A small output CFB will be nearly as tall as a much higher output unit. The design air, fuel residence time in the combustor is typically 3-5 seconds. With the nominal 20 feet per second velocities seen in the CFB's the boiler ends up being 60 to 100 feet tall regardless of unit size. Therefore, the CFB does not effectively scale down to the smaller unit sizes. Manufacturers feel that due to this scale-down problem, the smallest economically practical size unit is between 50,000 and 100,000 lb/hr of steam. The bed area of a CFB is less than one-half that of a BFB. However, due to the CFB's much taller height requirement the boiler volumes are nearly the same. The important fact is that the boiler heat transfer surface can be reduced for either BFB or CFB over that of a pulverized or stoker coal boiler due to the heat transfer rate of a fluidized bed combustor.

Plant designs were developed at 350 psi saturated for each technology in size ranges of (50,000; 100,000; and 200,000 lb/hr) using four different coals (Ohio 4% sulfur; Illinois 3% sulfur; Pennsylvania 2% sulfur; and West Virginia 1%). From these designs, plant capital and erection costs were estimated using vendor quotes for major equipment. Annual O&M costs were developed for each of the designs and are summarized on several curves which compare annual total cost per unit size, as well as cost per lb/hr of steam produced.

Levelized steam costs were calculated for various combinations of technology and include all costs associated with the production of steam over the life of the plant. Levelized steam costs represent an economic measure which allows direct comparison between alternatives. A number of curves have been developed which compare the various design options (BFB vs CFB, type of coal, etc.).

Total capital requirements at startup (5/1/88) were projected to range from \$176 per lb/hr for a 50,000 lb/hr sized boiler to \$113 per lb/hr for a 200,000 lb/hr sized boiler. The levelized cost per pound of steam varied depending on coal type, technology and unit size. The cost ranges for Illinois 3% S between 50,000 lbs/hr and 200,000 lbs/hr are compared for BFB and CFB and are provided in Table 1.

INDUSTRIAL AFBC CONCEPTS

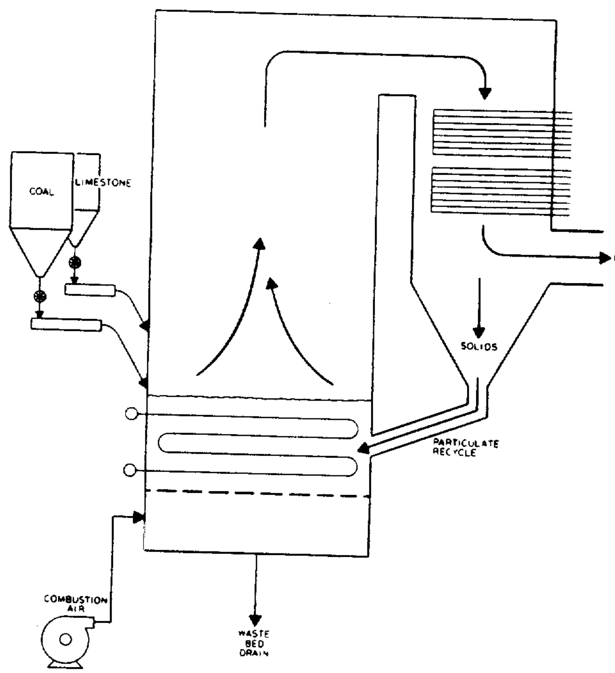
Bubbling Bed

The bubbling bed schematically shown in Figure 1 is characterized by the presence of a dense zone of granular particles supported by an air distributor

TABLE 1
MAXIMUM COST RANGES BETWEEN ALTERNATIVES

	Annual Costs (4/30/89) ($/1000 Lb Steam) Levelized (12%, 20 yrs)
Circulating Fluidized Bed	
50,000 lb/hr, Illinois 3% S	8.75
200,000 lb/hr, Illinois 3% S	6.22
Bubbling Fluidized Bed	
50,000 lb/hr, Illinois 3% S	9.12
200,000 lb/hr, Illinois 3% S	6.57

plate or grid. As combustion air enters the bed, it fluidizes these solids, i.e., causes them to assume a turbulent motion. Air velocities in the bed are kept below entrainment velocity and there is a well defined fluid bed surface.



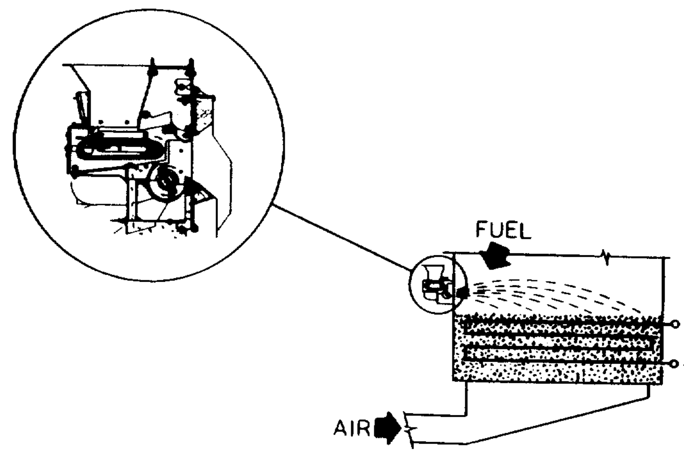
BUBBLING FLUIDIZED BED BOILER

FIGURE 1

Minimum and maximum allowable air velocities depend on solids density and sizing. These may vary from 1 to 12 ft/sec, however, normally the range is 4 to 8 ft/sec. The higher velocity is desirable to minimize bed size, but this reduces fuel and sorbent retention time so unburned carbon loss increases and sulfur capture decreases.

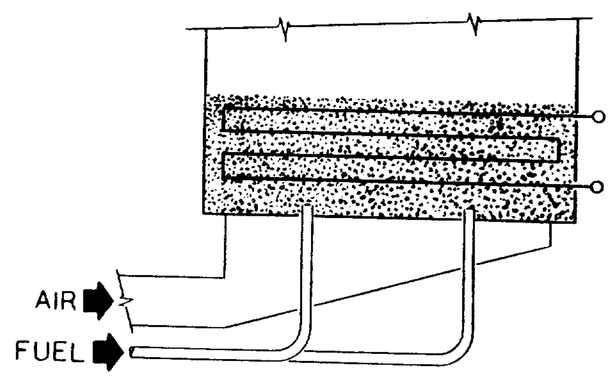
The furnace waterwall enclosure is inadequate with larger plan areas to absorb the heat required to maintain a 1450°-1650°F bed temperature. Therefore, heat must be extracted from the bed by submerging the heat transfer surface within the dense bed. In the smaller and low pressure units this is normally steam generating surface and may be arranged for natural circulation or a pumped circuit.

Two accepted methods of fuel (coal) feed to BFBs are over-bed and under-bed. The over-bed feeding system, which is the simplest, employs a mechanical or pneumatic spreader to distribute the fuel uniformly over the surface. A typical spreading device is shown in Figure 2. These feed types cover a long narrow strip so multiple feeds are needed for wide beds. A coarser fuel can be used with over-bed feeding, normally 3/4" x 0, with a 1/4" x 0 limestone bed. The under-bed feed system shown in Figure 3 requires a finer (1/8" to 1/4" x 0) grind coal. The under-bed feed system requires one feed point per 10-20 sq ft of bed area. This requires splitting the coal stream. In general, the sulfur capture is improved by under-bed fuel feed, as well as the carbon utilization and NO_x reductions.Carbon and sorbent utilization with either system can be improved by recycle. Recycle rates for the bubbling bed combustor are usually kept below about 4 lb of recycle solids per lb of coal feed; however, some European manufacturers have used higher values.



ABOVE BED FUEL FEED SYSTEM

FIGURE 2



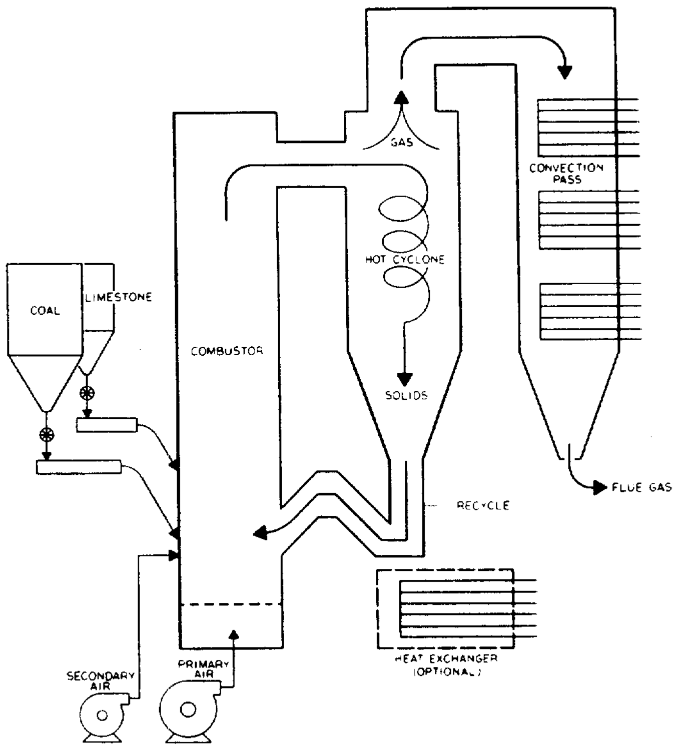
UNDER BED FUEL FEED SYSTEM

FIGURE 3

Circulating Bed

The circulating fluid bed schematically shown in Figure 4 uses an air distribution plate similar to the bubbling bed. However, it employs higher fluidizing air velocities, typically 15-30 ft/sec. Solids density in the zone immediately above the air distributor plate may approach those for a bubbling bed but gradually reduce over the combustor height. Therefore, there is no defined bed surfaces and a large part of the bed material is entrained and carried over to a solids separating device where the solids are captured and recycled. Recycle rates are 50 to 100 lb of solids per pound of coal feed. Some designs employ an external heat exchanger to cool the solids before recycle. Other designs employ only waterwalls in the combustion chamber to achieve cooling.

The heat transfer mechanism in both units is primarily from the burning fuel particle to the bed solids and from the solids to boiler surface. The presence of solids gives a much higher heat transfer coefficient than that achieved by flue gas. Although the solids-to-surface temperature differential in the furnace is lower than the flue gas to surface in a pulverized coal-fired unit, a higher overall rate of heat transfer can be obtained because of the solids contact with the heat transfer surface. Because of this higher heat transfer rate, a fluid bed boiler requires less heat transfer surface than a comparable size pulverized coal (PC) or stoker-fired boiler. The heat transfer for either type is similar to a PC or stoker unit in the convection pass.



CIRCULATING FLUIDIZED BED BOILER

FIGURE 4

DESIGN FEATURES

Bed Sizing

The cross-sectional area of a fluidized bed depends primarily on the following:

- o Thermal input required, Btu/hr
- o Firing rate (Btu/ft^2-hr) or superficial gas velocity (ft/sec)

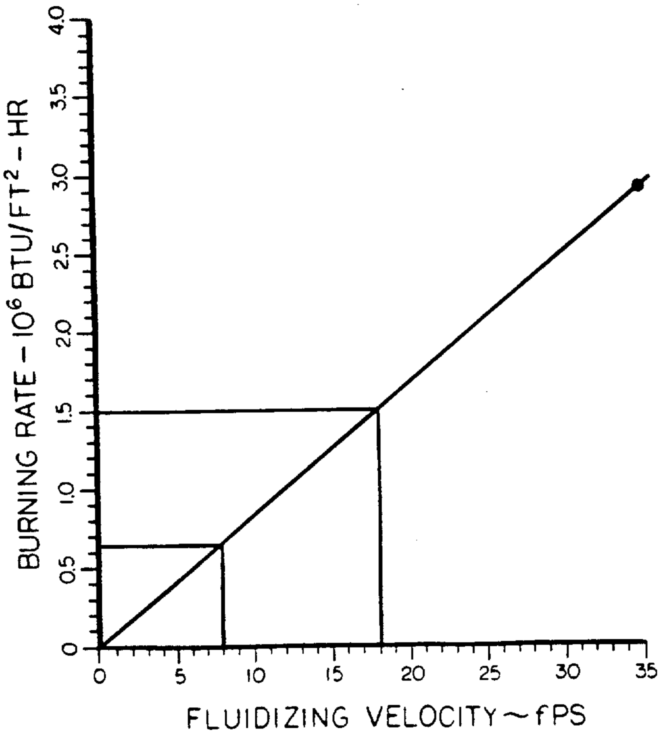
The amount of fuel which can be burned per sq ft of bed area is primarily dependent on gas velocity through the bed and, to a lesser extent, on the amount of excess air used and the operating temperature. In general, excess air is maintained in the 10 to 20 percent range for complete combustion and bed temperature in the 1500 to 1600°F range for good sulfur retention.

Assuming a 1550°F operating temperature and 15 percent excess air, the burning rate per sq ft of bed versus fluidizing velocity is approximately as shown in Figure 5.

The approximate bed plan area for the cases being considered, based on combustion calculations, are shown in Table 2.

Bed Temperature

The primary reason for controlling bed temperature is sulfur capture. With most limestones, the optimum temperature for sulfur capture is about 1550°F. The



BED AREA SIZING

FIGURE 5

percent capture for a given stoichiometric ratio drops significantly at ±100°F based on published data. Another advantage of lower combustion temperatures (compared to pulverized coal), is lower NO_x emissions. NO_x production begins to increase significantly above 1650°F. Carbon utilization is also affected by bed temperature and improves with increasing temperature. If the amount of unburned carbon is high, particulate capture and recycle can be used to improve efficiency. The upper limit on bed operating temperature is reached when ash agglomeration begins to occur.

The amount of heat release in a bed operating at normal excess air would raise the temperature to unacceptable levels without heat removal. For this reason, heat transfer surface is needed in a bubbling bed. A high solids recycle rate is employed to control bed temperature with circulating fluid beds. One method used to obtain a high rate is to control the quantity and distribution of the solids in the

TABLE 2
BED PLAN AREA

TYPE - Bubbling Bed

Fluidizing Velocity - 8 feet per second (fps)			
Steam Flow Lb/Hr	50,000	100,000	200,000
Fuel Input 10^6 Btu/Hr	58	117	234*
Bed Area Ft^2	92	180	360
Dimensions	8 x 12	10 x 18	15 x 24

TYPE - Circulating

Fluidizing Velocity - 18 fps			
Steam Flow Lb/Hr	50,000	100,000	200,000
Fuel Input 10^6 Btu/Hr	57	113	227*
Bed Area Ft^2	39	76	152
Dimensions	6.5 x 6.5	9 x 9	10 x 15

*Combustion efficiencies used are averages of data supplied by manufacturer.

combustion chamber, thus controlling rate of heat transfer to the walls. A second method is to cool the recycle solids using an external bubbling bed heat exchanger. Both systems are being commercially offered by the manufacturers. The use of preheated combustion air tends to raise bed temperature, which requires greater bed cooling.

Air Distribution

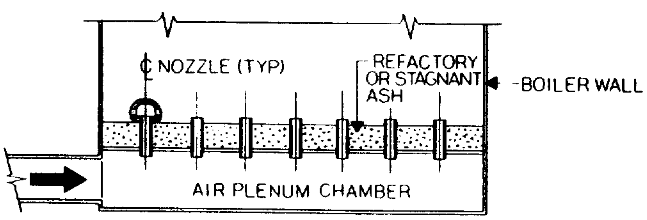
The air distributor has several functions and should be designed with the following characteristics:

- Support for the bed
- Provide uniform air distribution
- Promote particle movement
- Minimize pressure drop for adequate distribution
- Operate over long periods without problems

There are a number of configurations used, the more common being a plate construction with air nozzles.

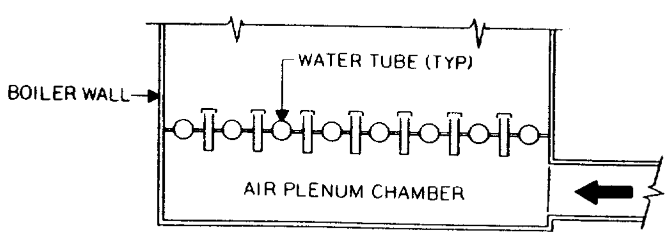
The high heat transfer coefficients in the bottom of the dense bed combustion zone and the 1600°F temperatures next to the distributor plate require thermal protection for the distributor plate. This is accomplished either by insulation or cooling.

Two common approaches to grid plate insulation are the use of a refractory layer or a stagnant zone of ash. These are depicted in Figures 6 and 7.



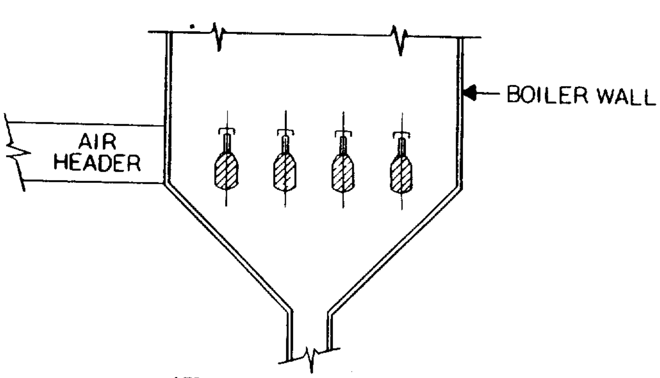
INSULATED AIR DISTRIBUTOR PLATE

FIGURE 6



WATER COOLED AIR DISTRIBUTOR PLATE

FIGURE 7



AIR HEADER DISTRIBUTOR SYSTEM

FIGURE 8

Some form of ash drain from the distributor plate is usually provided. An air distributor, utilizing an air header, is often used instead of a solid distributor plate when burning a high ash material or a material containing coarse inerts (rocks). A typical design is shown in Figure 8. This system allows large rocks to pass through the air distributor to discharge. Any of these designs can be used for either the CFB or the BFB.

Each of the distributor systems can employ a variety of nozzle designs. The nozzles serve to create a pressure drop to obtain uniform air distribution. In addition, by extending above the distributor plate the nozzles can shield the plate from high temperature bed material. Distributor systems must also be designed to prevent drainage of bed material into the air plenum during shutdowns.

Bed Pressure Drop

In addition to the pressure drop across the air distributor there is an additional pressure drop created by the bed material. In the fully fluidized state, this pressure drop is equal to the weight of the bed material (bed inventory) for either the bubbling or entrained type. For the bubbling bed, this occurs fairly uniformly over the bed depth. For the entrained bed, the pressure drop per unit of height is usually higher in the lower regions, indicating a higher solids loading. The measured pressure drops over the height of an entrained bed can be used to calculate the solids distribution for controlling furnace wall heat absorption. The pressure drop can also be used to measure the amount of bed material and to activate a bed drain system.

Fuel and Limestone Feed System

The fuel (coal) is expected to be delivered to the plant site by truck or rail and placed in storage. Coal can either be received crushed and presized or it can be received unprocessed. If the coal is already crushed, it is normally stored in closed silos to prevent blowing particles. In some cases, these silos may be directly above the boiler fuel feed system and serve as live storage. In other cases, where larger volumes are required, these storage silos may be remote and the coal transferred as required to the boiler feed bins.

Unprocessed or raw coal is normally placed in ground storage piles. Coal is recovered from this pile, goes to a prep plant where it is crushed to the desired size and conveyed to bin(s) serving the fluidized bed boiler. The extent of crushing will depend on the type and manufacturer of the fluidized bed. Normal sizing will range from a coarse end of 3/4 in. x 0 to a fine end of 1/8 in. x 0. If an under-bed feed system is used, the coal processing plant will usually include coal drying equipment. For coal to feed properly in pneumatic under-bed systems it is typically limited to 1% to 3% surface moisture.

Coal from the working bin will normally be removed by some form of variable rate metering type feeder(s). In most cases fuel will pass through some form of air seal (rotary valve) before entering a pressurized pneumatic line conveying it to the fluidized bed boiler. For some types of fluidized beds, this air-fuel stream may be split for multi-nozzle injection.

The limestone usually involves smaller quantities than the fuel but typically will be handled in a similar manner. Typical limestone use will range from 1/4 in. x 0 to a pulverized material. For cases where usage is small, the limestone may be purchased as a precrushed material.

Sulfur Capture

Sulfur or SO_2 capture in a fluidized bed combustor is usually achieved by the addition of limestone or dolomitic stone. Experimental work has been done with other materials such as alkalis and rare earths. In most cases, these are pelletized and employ a regenerative step.

The basic reactions are:

$CaCO_3 \rightarrow CaO + CO_2$ — Heat required 76,590 Btu/lb-mole (Calcination)

$CaO + SO_2 + 1/2\ O_2 \rightarrow CaSO_4$ — Heat released 215,172 Btu/lb-mole (Sulfidation)

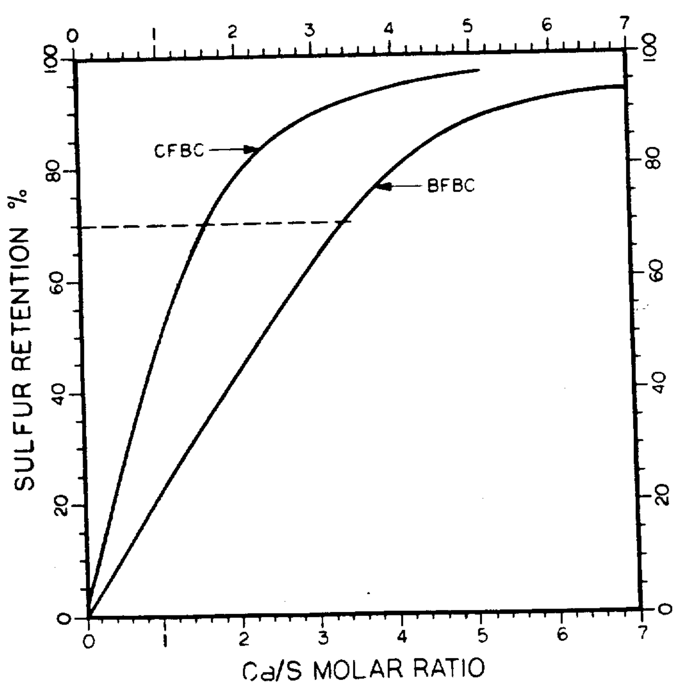
The optimum temperature for the sulfidation reaction is about 1550°F. In most cases, an excess of limestone is needed to achieve the desired SO_2 capture. Only a portion of calcined limestone is sulfated and therefore the net overall loss or gain of Btu is small in most cases.

The bubbling fluidized bed combustor normally employs about a 1/4 in. x 0 grind limestone which constitutes the major portion (more than 90%) of the bed material. Sulfidation is a surface process so only a small percentage of coarser particles react. Bed agitation removes some of the particle coating and exposes new surface.

The circulating fluidized bed combustors tend to use a fine grind limestone and employ greater agitation, thereby achieving better limestone utilization than bubbling beds. The CFB combustors also recirculate bed material 50 to 100 times which provides a very long residence time and aids limestone utilization.

There are a variety of tests which aid in proper limestone selection. These include elemental composition, porosity, density, pore size distribution, hardness and thermogravimetric analyses tests (TGA). While these are useful as a guide, the ultimate method is to conduct pilot plant tests with the fuel and limestone selected for use.

The plot in Figure 9 shows the predicted percentage of sulfur removal that might be expected for a good limestone grade and compared limestone usage for both BFB and CFB.



PROJECTED PERFORMANCE OF SORBENT

FIGURE 9

Carbon Utilization

Operating a fluidized bed at elevated temperatures favors higher carbon utilization.

However, the need to capture sulfur with limestone limits upper bed temperature to about 1600°F. The initial bubbling fluidized beds were operated without fly ash recycle, however, many experienced unburned carbon losses in excess of 10-15 percent. By installing mechanical dust collectors and employing recycle unburned carbon losses under 5 percent can be attained.

The circulating fluidized bed on the other hand includes an exit solids removal and recycle system as a basic part of its design concept. The carbon utilization of a circulating bed tends to exceed that obtained with bubbling beds; however, some applications have now resorted to a secondary dust collection and recycle system to further reduce unburned carbon losses in CFB's to less than 1%.

NOx (Nitric Oxide). NO_x emissions from a combustion process result from two sources: (1) the conversion of nitrogen from the combustion air, and (2) the conversion of fuel-bound nitrogen. It is generally accepted that NO_x (primarily NO) is formed by the union of atomic nitrogen and oxygen and not from the molecular species. Thus thermal NO is most readily formed at the higher temperatures when the N = N and O = O bonds have been broken. The fuel-bound nitrogen is usually associated with an organic molecule and passes through the atomic form when released. If an oxygen atom is available, NO is formed.

The lower combustion temperatures in a fluidized bed tend to produce less thermal NO_x. By operating the lower zone in a fuel rich manner and completing combustion by overfire air, the fuel-bound nitrogen has limited access to an oxygen atom and tends to form N_2. In general, there is minimal NO_x formation from disassociation of nitrogen compounds. As NO exits a combustion process it reacts with atmospheric oxygen to form NO_2. The CFB generally controls NOx better than a BFB due to staged combustion which is typical of the CFB design.

CO (Carbon Monoxide). Carbon monoxide is formed by two basic reactions,

$$C + 1/2\ O_2 \rightarrow CO \quad \text{or} \quad CO_2 + C \rightarrow 2\ CO$$

Both are considered the result of incomplete combustion, i.e., a lack of sufficient excess air or the presence of excess carbon. Although combustion processes are generally operated with an overall excess of oxygen in the final flue gas, most have some zones where excess carbon exists due to incomplete mixing or staging. Although both types of fluidized beds employ a high degree of turbulence, each tends to show emissions of CO ranging from 50 ppm upward. The resultant loss of boiler efficiency is negligible.

Hydrocarbon Emissions. Hydrocarbon emissions are thought to result from incomplete combustion reactions similar to CO. The range of hydrocarbons which can be emitted is fairly vast and are of the form C_NH_{XN} where N can range from 1 to 36 and X from slightly less than 1 to 4 (methane CH_4). Hydrocarbon emissions are often divided into two classes, methane and non-methane. Although most air districts have not set limits on hydrocarbon emissions it is of growing concern. Very little data exist on fluidized bed emission of these gases.

Sulfur Dioxide (SO_2). Sulfur dioxide emissions are formed during the combustion process from the combination of sulfur in the coal and oxygen in the combustion air. The amount of SO_2 produced and emitted depends on the sulfur content of the coal and is normally calculated on a pounds per million Btu basis. As has been previously stated, sulfur capture is easily accomplished in an AFBC by the addition of a sulfur sorbent material into the combustion process. The normal sorbent used is limestone. The amount of sulfur capture or the percent removal is dependent on the amount of sorbent used and its reactivity rate. All current and projected emission standards can be met. Current regulations are 1.2 lb/MBtu. With sufficient addition of sorbent, nearly all the SO_2 can be removed.

Particulate. Particulate emissions from an AFBC consist of fly ash, calcined limestone and gypsum. Much of the solid wastes produced by an AFBC are carried out by the flue gas. Currently, the best available accepted technology is the application of a fabric filter baghouse system. Baghouse inlet loadings are typically 10 grains per actual cubic foot and outlet levels are controlled to 0.03 lb/MBtu fuel fired.

Heat Transfer. The heat transfer in an FBC is primarily by a convection-conduction mode due to both particle and gas contact with the tube surface. Some radiative transfer also occurs, but the primary variable occurring is between lean-phase and dense-phase. As the solids loading varies from 0 to about 50 lb/ft^3, the heat transfer coefficient goes from about 10 to 60 Btu/ft^2-Hr-°F. Dense-phase transfer takes place in a bubbling bed; lean-phase transfer occurs in the freeboard of a bubbling bed and in the convection passes of both type units.

The heat transfer to the cooled wall surfaces of a circulating fluid bed falls in an intermediate zone between lean and dense. The solids loading of the gas in the combustion is controlled both by internal and external recirculation. Some of the vendors rely on the control of solids to control heat transfer in the combustion zone using combinations of split air admission and external solids recycle rate.

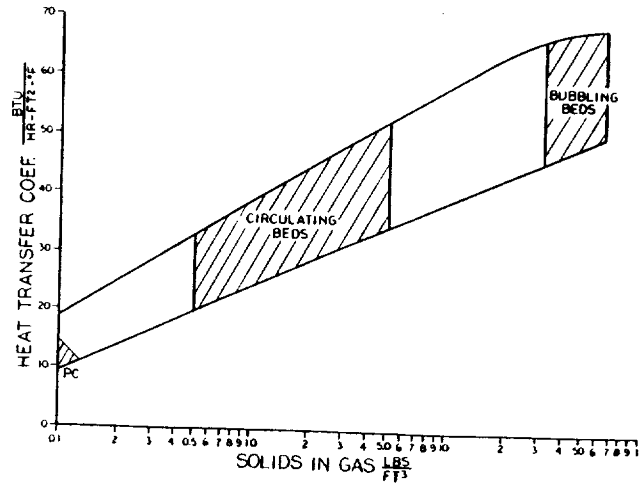
Figure 10 is a graphic representation of how the heat transfer coefficient is expected to vary with solids loading of the gas stream. The lower values are those typically obtained in stoker or pulverized firing where the solids loading of the gas stream is near zero or fairly low. The upper values are those obtained for "dense" or bubbling bed surface and external heat exchangers. The lower values in this area are typical of inbed surface in a fired unit. The values shown for a circulating bed are based on very limited data. The lower values shown represent the upper furnace zone prior to the inlet to the solids separator. The higher values represent the lower furnace zone above the refractory. The upper solids loading limit is set by bed pressure drop. One method of controlling furnace heat absorption is controlling the solids loading distribution over the furnace height of a circulating fluidized bed using primary/secondary air split. The spread in values partially occurs as a result of the particle sizing and, to some extent, the surface arrangement.

Capital Costs

Capital costs were estimated by combining historical information with process equipment cost data from vendors. Order of magnitude cost estimates for steam plants were developed from process equipment cost estimates. Cost relationships were calculated between construction costs and the purchase price of the process equipment.

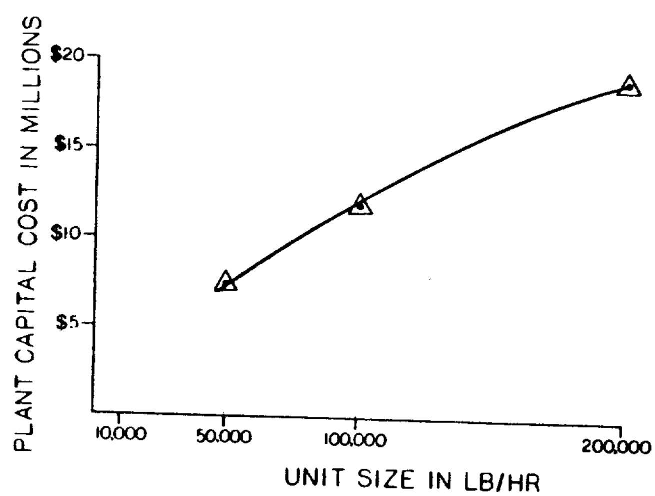
A model for the steam generation plant was prepared using the historical relationships for industrial power and steam generation. This model initially calculated the installed cost of the process equipment. The costs for the other direct accounts were then calculated as percentages of the process equipment accounts and proportioned to individual cost elements.

The capital cost differences between BFB and CFB was investigated and, it was discovered that within the level of accuracy of the budget pricing there would be little cost difference between the circulating bed and bubbling bed combustor for unit sizes above 50,000 lb/hr. Manufacturers did not feel they could economically justify CFB units less than 50,000 lb/hr, and some manufacturers felt the breakpoint was about 100,000 lb/hr. The costs provided represent the costs for a steam generating facility using either type of combustor.



EFFECT OF SOLIDS LOADING ON HEAT TRANSFER RATE

FIGURE 10



TOTAL PLANT CAPITAL COSTS

FIGURE 11

Figure 11 shows the total plant capital cost in millions of dollars based on unit size in lb/hr. The cost increases from $7.5 million for the 50,000 lb/hr unit to $19.0 million for the 200,000 lb/hr unit.

The combined effect of escalation and interest during construction may add 14 to 19 percent of the total plant cost to the final total capital. These values are listed in Table 3. The difference in cost for the same unit size reflected in the coal type is a result of the different requirements for the material handling systems.

Operation and Maintenance Costs

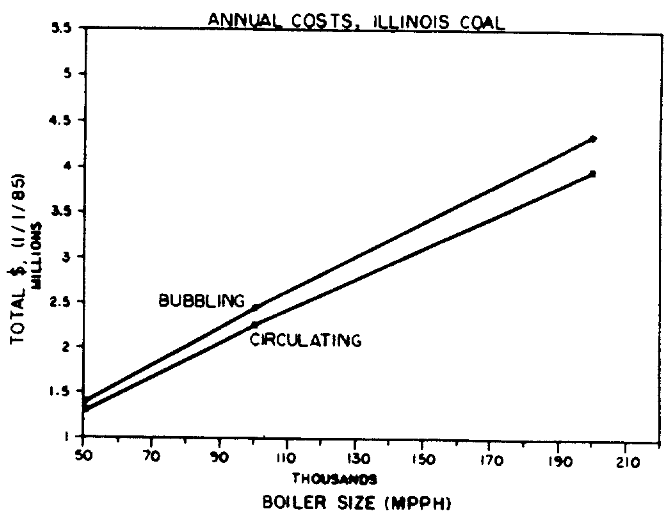
Variable costs can be categorized into two major components: fixed operation and maintenance costs and variable operating costs. The fixed operation and maintenance costs are those which do not vary with the production rate of the facility but are affected by escalation rates. In this analysis, operating and maintenance labor and maintenance parts were considered to be fixed O&M.

The variable operating costs are those which are a direct function of both the production rate and the escalation rates. Coal, limestone, ash, raw water, auxiliary power, water treatment and natural gas start-up costs were included in the analysis.

Total O&M costs as function of technology and size are illustrated in Figure 12. The most significant difference between the two technologies is the increased limestone cost for the BFB, which is nearly twice that of the CFB. This is due to the BFB's higher limestone useage rate for the same sulfur removal efficiency as shown in Figure 9. There is also a slight increase in coal cost for the BFB because of reduced boiler efficiency.

TABLE 3
CAPITAL COSTS ($/Lb/Hr) BY BOILER SIZE AND COAL TYPE

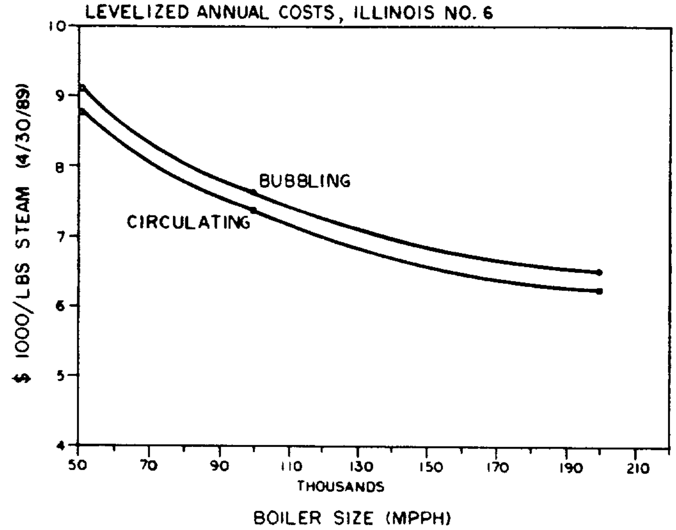
Boiler Size (lb/hr)	Coal Type	Total Plant Cost (1/1/86)	Escalation and Interest During Construction	Total Capital Requirement (5/1/88)
50,000	Ohio	152	25	177
	Illinois	151	25	176
	Pennsylvania	149	24	174
	West Virginia	149	24	173
100,000	Ohio	120	20	140
	Illinois	119	20	139
	Pennsylvania	119	20	139
	West Virginia	117	20	137
200,000	Ohio	95	18	113
	Illinois	94	18	112
	Pennsylvania	93	18	111
	West Virginia	93	18	111



O&M COST COMPARISION: TECHNOLOGY & SIZE
FIGURE 12

Levelized Cost Results

The most meaningful comparison of alternative technologies, sizes and fuels for this analysis is that of levelized steam costs. These are defined as levelized annual costs divided by average annual pounds of steam produced. This measure represents an average cost of steam which reflects all of the costs projected to be incurred by the plant over its service life.



TECHNOLOGY COMPARISION: SIZE
FIGURE 13

Levelized steam costs for all combinations of technologies, sizes and fuels are listed in Table 4. The CFB was projected to be the least expensive in all cases. However, the percent difference was relatively small and ranged from approximately 1 percent for the 50,000 lb/hr West Virginia coal to approximately 7 percent for the 200,000 lb/hr high sulfur Ohio coal.

TABLE 4
LEVELIZED STEAM COSTS BY TECHNOLOGY, BOILER SIZE AND FUEL TYPE

Boiler Size (lb/hr)	Coal Type	Levelized Steam Cost ($1000 Lb/Yr, 4/30/89) CFB	BFB	% Difference
50,000	Ohio	8.61	9.07	5.3
	Illinois	8.75	9.12	4.2
	Pennsylvania	8.45	8.68	2.8
	West Virginia	8.39	8.50	1.3
100,000	Ohio	7.17	7.62	6.2
	Illinois	7.32	7.67	4.8
	Pennsylvania	7.05	7.26	3.0
	West Virginia	6.95	7.06	1.6
200,000	Ohio	6.08	6.53	7.3
	Illinois	6.22	6.57	5.6
	Pennsylvania	5.94	6.16	3.6
	West Virginia	5.88	5.99	1.8

The decline of levelized steam cost with increasing plant size is illustrated in Figure 13 for the Illinois coal. The CFB costs are slightly less than the BFB costs for all size ranges examined. This is due to combustion efficiency and limestone utilization.

SUMMARY

Both BFB and CFB offer distinct advantages over traditional pulverized or stoker coal boilers. The technical viability of both technologies has been proven by full scale operating units and general acceptance of the fluidized bed concept is growing. Both designs allow the use of low grade, hard to burn fuels without the need for expensive post combustion scrubbing equipment. However, as illustrated in this paper, there are differences between BFB and CFB in cost, operation, and application.

AFBC is not new. In fact, AFBC was first studied and used in the 30's and 40's. Then in the 60's and 70's BFB was commercially developed and tested. In the 80's, CFB was developed as a result of desired increases in performance and efficiency. CFB has proven itself superior to BFB in many ways. In addition to CFB's better performance in combustion efficiency, sulfur capture is significantly greater with less limestone required. Its simpler fuel feed system and lack of in-bed tubes means lower maintenance and operating costs. A CFB's biggest disadvantage is its higher cost for smaller units and its relatively large vertical height requirement.

A BFB offers some advantages in certain cases and, therefore, must not be forgotten. Since CFB is not cost-effective for very small units (below 50,000 lbs/hr), BFB is the practical choice in this size range. CFB units up to 1.0 million are currently being built and the major manufacturers are confident of their capability to produce a 1.5 million lb/hr CFB unit. However, above 1.5 million lbs/hr, scale-up limitations make technical feasibility questionable. Therefore, 2 to 3 million lbs/hr units would have to be BFB's, which are currently considered to be technically feasible. Other applications where BFB may be a better choice is for waste incineration and in conversion of pulverized coal (PC) units to fluidized bed combustion. The choice for waste incineration is dependent upon the waste type and application. Recently, growing emphasis on plant life extension and SO_2 control has produced increased interest in BFB retrofit of older PC units as exemplified by Northern States Power's Black Dog unit and Montana Dakota Utilities.

A summary comparison of BFB and CFB is shown in Tables 5 and 6.

TABLE 5
BFB/CFB COMPARISON

	BFB	CFB
SIZE	ALL SIZES	100,000 thru 1,500,000 #/hr
Combustion Eff	Base	2-3% Better
Sorbent Use	Up to 100% More	Base
Bed Area	2-1/2 Times	Base
Comb Volume	Similar	
Heat Transfer	Average Similar	
Fuel Feed	Overbed/-Underbed	Inbed
Foot Print	BFB More Above 250,000 #/hr	CBD More Up to 250,000 #/hr CFB Does not Scale Down Well
Heat Recovery	Inbed Tubes	No Inbed Tubes
Boiler Controls	Conventional	Conventional
Material Handling	Conventional	Conventional
Steam/Water	Conventional	Conventional
Startup	4 Hours	8 Hours
Cost	Similar	
Plant Horse Power Auxiliary Load	Similar if BFB is overbed feed system, BFB more if under bed feed system	
O&M	Base	Lower

TABLE 6
BFB/CFB COMPARISON
100,000 Lb/Hr. Unit

	BFB	CFB
Height	40'	100'
Length	35'	55'
Width	20'	30'
Bed Area	180 Ft^2	81 Ft^2
Combustion Vol	6300 Ft^3	6000 Ft^3
PERFORMANCE		
Boiler Eff.	87.1%	88.5%
Ca/S (70% Ren)	3.2:1	1.5:1